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ACTIVATED SLUDGE TREATMENT WITH OXYGEN

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Robert A. Taft Water Research Center
Cincinnati, Ohio

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By

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ABSTRACT

An oxygen activated sludge system, consisting of a unique gas-tight biological reactor with less than 2.5 hours of aeration time and gravity clarification, produces effluents comparable in BOD and suspended solids to those of a parallel diffused aeration system with an aeration time of 3.6 hours. Consistently over 95% of the supplied oxygen was utilized in the reactor which employed co-current liquid and gas flow contacting.

The mixed liquor suspended solids were maintained between 4,000 and 6,000 mg./l. The mixed liquor increased to over 8,000 mg./l. when operated with alum addition for phosphorus removal. The underflow solids concentrations were between 1.00% and 1.40% without alum and 2.14% during the alum addition period.

The microbial organisms visually were essentially the same as a typical conventional system however, their rate of activity was greater above an SRT of 6 days. The total production of solids was significantly less than the similarly operated diffused air system above an SRT of 6 days. As little as 0.35 lb. of solids/lb. of BOD were produced at an SRT of 13 days. Filamentous growth was observed below an SRT of 5 days but not above.

In warmer temperature periods, peak overflow rates of 1,940 gal./(day)(sq. ft) were observed. In the colder temperature periods, overflow rates of 975 gal./(day)(sq. ft.) were the maximum obtained during the first year's operation.

Activated Sludge Treatment With Oxygen*

By

John B. Stamberg, Dolloff F. Bishop, and Gordon Kumke**

The use of pure oxygen within the activated sludge process dates back some 20 years. Pirnie (1) proposed a system of predissolving pure oxygen in high concentrations in the influent wastewater before entering a non-aerated mixed reactor. The process was termed "a bio precipitation system." Biological success was achieved by Okun (2) in bench scale tests and later by Budd and Lambeth (3) on a pilot scale, but oxygen utilization efficiencies of 20-25% were too costly. Okun and Lynn((4) and later Okun (5) showed an increase in the effective sludge activity in the mixed liquor by reducing or eliminating anaerobic periods such as can occur in clarification.

McKinney and Pfeffer (6) more recently reviewed the use of oxygen in activated sludge. Increased metabolism rates, produced by eliminating periods of zero dissolved oxygen, would increase treatment efficiencies in overloaded plants and reduce the size required for new plants. Thus, potential reductions in capital investment were viewed possible for oxygen systems.

Union Carbide recently developed the UNOX System (7) which is an oxygen-aeration-activated sludge system with an oxygen utilization of over 90%. This oxygen-activated sludge process (UNOX) is presently being piloted in several

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locations. The Environmental Protection Agency, Washington, D. C. Pilot Plant has been evaluating the oxygen-activated sludge process since May of 1970. The objectives of the study is to determine process performance and operating requirements on the D.C. primary wastewater, a characteristically weak wastewater.

Oxygen Pilot System

The oxygen-activated sludge process in the EPA-DC Pilot Plant employs a unique gas-tight biological reactor with a conventional gravity clarifier (Figure 1). In the EPA-DC Pilot Plant, primary effluent from the District of Columbia's plant is fed to the oxygen reactor either on steady state flow or on a predetermined daily cycle or diurnal variation, normally with a 2.3:1 (45 - 105 gal./min.) daily flow variation (Figure 2).

The 8,100 gallons reactor provides 1.95 hours of retention time at the nominal influent flow of 100,000 gal./day. At the peak daily flow, the retention time is 1.28 hours.

The reactor is sealed to prevent loss of oxygen and includes submerged hydraulic entrances and exists as well as simple water-sealed mixing equipment. It is divided into stages to provide the proper tank geometry for efficient mixing and also for efficient oxygen usage.

The efficient oxygen usage is achieved by co-current contacting of the mixed liquor and oxygen gas through the various stages. The addition of pure oxygen to the reactor is controlled by a pressure regulator. An inlet oxygen control valve actuated by the pressure regulator maintains the overhead gas at a selected pressure usually between 1" and 4" of water. Even with large instantaneous fluctuations in oxygen consumption, the oxygen control valve easily maintains the selected pressure. The overhead gas pressure is normally selected to produce an oxygen concentration at approximately 50% in the exhaust

gas from the reactor. Pure oxygen is introduced to the first stage where the peak oxygen demand occurs. As the oxygen is used in biological metabolism, the respirated carbon dioxide reduces the oxygen concentration in the overhead gas flowing co-currently with the mixed liquor through the succeeding stages. The successive decrease of both oxygen supply and demand produces efficient oxygen use before the residual gas is exhausted from the reactor.

The dissolved oxygen level in the mixed liquor is controlled between 4.0 and 8.0 mg./l. by adjusting the recirculation rate of the oxygen gas within the individual stage. The compressor in each stage pumps the overhead gas through the rotating mixer impeller to provide efficient dispersion and mixing of the recirculated gas. The recirculation rate in each stage may be set either manually on the basis of a dissolved oxygen analysis or automatically using a control system with a dissolved oxygen probe as a sensor.

Because the high mixed liquor solids fouled the peripheral feeding orifices, the original peripheral fed gravity clarifier was later converted to a center feed clarifier (Figure 3). As a center feed clarifier, the cross-sectional area of the circular tank is 107 sq. ft. below the peripheral feeding skirt, 96 sq. ft. above the skirt, and 78 sq. ft. between the center feed well and the weir. At nominal flow of 70 gal./min., the overflow rate is 940 gal./(day)(sq.ft.) below the skirt, 1,050 gal./(day)(sq.ft.) above the skirt, and 1,280 gal./(day)(sq.ft.) between the well and the weir. At peak flow of 105 gal./min., the rates are 1,410 gal./(day)(sq.ft.) below the skirt, 1580 gal./(day)(sq.ft.) above the skirt, and 1,940 gal./(day)(sq.ft.) between the well and the weir.

Process Operations

The system was started in May by gradually increasing the flow as the

mixed liquor solids increased. In two weeks, the mixed liquor suspended solids (MLSS) had increased from 700 mg./l. to 6,000 mg./l. as influent flow increased to 65 gal./min. The sludge volume index (SVI) was 35. Unfortunately, on May 31, an operator left a recycle sample valve open and the mixed liquor was reduced to 1,500 mg./l. In an attempt to rebuild the mixed liquor solids at a 65 gal./min. flow rate, the biomass developed heavy filamentous growth and bulking of solids occurred (SVI over 300). After ten days of operation without change in the solids level or filamentous growth pattern, the flow was reduced to 30 gal./min. In several days, a settleable biomass developed (SVI=100) and the flow was then increased to 50-55 gal./min. for the remainder of June (Table 1 & 2, Period 1).

In July, the steady state flow was increased to 80 gal./min. with a detention time of 1.66 hours (Table1&2, Periods 2 &3). The 80 gal./min. operation continued through August 25, 1970. In August, the recycle rate and solids wasting were gradually reduced to increase the solids retention time (SRT). On August 16, 1970, the peripheral clarifier was converted to center feed (Figure 3). After August 25, the system was operated (Table 1 & 2, Period 4) on a diurnal pattern (Figure 2) expected for dry weather flow of the District of Columbia's future plant.

The solids retention time is the reciprocal of the biological growth rate. It is average retention time of the micro-organisms in the system (Equation 1).

$$SRT = \frac{\text{lb. Volatile Suspended Solids Under Aeration}}{(\text{lb. Volatile Solids Wasted} + \text{Volatile Solids in the Effluent})/\text{Day}} \quad \text{(Equation 1)}$$

In October (Table 1 & 2, Period 5) alum was added to the fourth-stage reactor to remove phosphorus by mineral addition (8). Increasing alum dosages were employed until poorly bioflocculated mixed liquor developed on October 25 and alum addition was discontinued.

During the colder weather, three operating modes were tried to redevelop

a mixed liquor with settling qualities comparable to the earlier operation. The modes were operating with 6,350 mg./l. of MLSS and a 5.5 day SRT (Table 1 & 2, Period 6), with 5,300 mg./l. of MLSS and a 13-day SRT (Table 1 & 2, Period 7), and with 3,940 mg./l. of MLSS and 4.7 SRT (Table 1 & 2, Period 8*).

Analytical Procedures

To evaluate the oxygen-activated sludge process, appropriate samples were manually composited over a 24-hour period. Samples were stored at 3°C to minimize biological activity.

The 5-day biological oxygen demand (BOD) of the composite samples was determined by the probe method (9); the ammonia (9) and nitrate-nitrite (10) on a Technicon Automatic Analyzer. The total organic carbon (TOC) (11) was measured on a Beckman Carbonaceous Analyzer. The total phosphorus (12) was determined by the persulfate method. All other analyses employed Standard Methods (13). Soluble phosphorus and soluble BOD were filtered through a standard glass suspended solids filter before analyses.

Carbon, Nitrogen and Phosphorus Removal

During the various tests periods (Table 3), the oxygen aeration system produced a good quality secondary effluent. The reduction of biological oxygen demand (BOD) from the primary effluent varied between 91 and 76%. The residual BOD in the effluent varied between 11 and 32 mg./l. The soluble residual BOD averaged less than 3 mg./l. in all the test periods, and indicated essentially complete insolubilization of BOD. Most of the residual BOD in the effluent was associated with the suspended solids and the overall removals were a function of clarification efficiency.

The effluent suspended solids varied between 24 and 58 mg./l. The removal efficiencies from primary effluent were as low as 36% in periods of poor clarification, and as high as 78% during periods of good clarification.

The chemical oxygen demand (COD) removal from primary effluent varied

between 61 and 82% with residuals between 45 and 99 mg./l. The TOC removal from primary effluent varied between 86 and 63% with residuals between 14 and 26 mg./l.

Nitrification started in late June after the elimination of the filamentous growth earlier in the month. As the flow was increased from 50-55 gal./min. to 80 gal./min., nitrification decreased (Figure 4). In August the solids wasting was decreased, increasing the SRT from 7.7 to 11.8 days. The nitrifying organisms then propogated faster than they were wasted and nitrification was reestablished. As in a parallel step aeration process, the nitrification decreased in the fall because of the metabolic changes in colder water. The nitrogen removal from the primary effluent during periods of nitrification and partial denitrification was as high as 39% (Table 4). The nitrogen removal decreased to as low as 9% during periods without nitrification.

Without alum addition, the phosphorus was removed through metabolic uptake and by the wasting of solids, thus, the removals varied with the metabolism of the mixed liquor. In September, with an SRT of 10.7 days (highly endogeneous metabolism), the total phosphorus removal was only 15%. The ratio of TOC to phosphorus was 100:5.3 in the waste sludge (1.4% solids). At lower SRT's with less endogenous respiration as in June and July, the phosphorus removals increased to 20%. The ratio of TOC to phosphorus in the wasted sludge increased to 100:6.6 and larger volumes of sludge were wasted.

In October, with alum addition (Figure 5), the phosphorus removal increased as the alum weight ratio (Al^{+++}:P) increased. For a dosage equal to an Al^{+++}:P ratio of 1.4:1, eighty percent of the phosphorus was removed to an average residual of 1.8 mg./l. as P and only a slight decrease in wastewater alkalinity and pH occurred. The filtered effluent (through 0.45 u.) contained an average of 1.6 mg./l. of "soluble" P. When the dosage was increased to a

ratio of 1.85:1 Al^{+++} to P, the residual phosphorus decreased to 0.62 mg./l. as P with 0.53 mg./l. in filtered effluent(through 0.45 u.) At this dosage, however, the average pH of the mixed liquor decreased from 6.5 to 6.0 and the biomass dispersed. Thus, alum addition was stopped to allow the mixed liquor to recover.

Biomass Characteristics

The mixed liquor suspended solids were similar visually to the micro organisms in conventional activated sludge. The mixed liquor biota was normally very well bioflocculated with active stalked ciliates growing on the bacterial mass. Zooflagellates and free swimming ciliates, although few in number, remained adjacent to or within the flocculated particles. Several varieties of large active rotifiers were plentiful. A few nematodes existed in the sludge. Normally, filamentous growth was not apparent. There was almost a complete absence of fragmented debris or unflocculated bacteria between the discrete particles (Figure 6).

When the endogenous respiration in the mixed liquor biota decreased, as in the winter at lower metabolic rates, the particles lacked color and were almost clear. As seen under a microscipe, the particles in mixed liquors with moderate SRTs (4-6 days in the winter) were large and well bioflocculated. Visually, the movement of the protozoa impart more movement to the particles than to the particles of oxygen sludge with a higher SRT or to particles of a conventional sludge. However, density of the particles were not measured.

While the system was not toxic to the nitrifying organisms, nitrosomonas and nitrobacter, the recirculation of respirated carbon dioxide within the reactor stages lowered the wastewater pH from 7.0 - 6.8 in the first stage to 6.4 - 61. in the final stage. With a system pH of approximately 6.5, the oxygen process more slowly established a nitrifying population than the conventional activated

sludge system operated at pH of 7.0 - 7.4. However, when the solids wasting was reduced to a level that the organisms propagated faster than they were removed, the population increased and substantial nitrification occurred.

While the oxygen activated sludge system on the District of Columbia wastewater was not free of filamentous growth in the SRT range less than 5 days, filamentous growth did not occur during operation above an SRT of 5 days. February, after maintaining the system at an SRT less than 5 days for approximately a month, filamentous growth, as in step aeration for similar SRT, began to propagate in the mixed liquor. A reduction of flow to increase the SRT reestablished a filamentous free sludge in several days.

With alum addition, the biological floc was uniformly round and of a more uniformed size than the floc without alum addition. Changes in the characteristics of the protozoa were not observed. With the alum, the volatile portion of the solids dropped to 67%.

On the District of Columbia wastewater, the oxygen process showed a much higher activity in the volatile portion of the solids (F/M ratio) for the SRT range above 6 days than the conventional aeration process (Figure 7). The F/M ratio is the ratio of the BOD applied to the mixed liquor volatile suspended solids (MLVSS) under aeration. Thus with oxygen, shorter retention times and lower solids production than in step aeration were achieved by operating the reactor at high SRT. The most important reason for the increased activity is that the dissolved oxygen is maintained between 4 and 8 mg./l. The independently controlled mixing also minimized sludge pockets, dead spots and the shearing of the floc particles and the mixed liquor entering the clarifier had a high dissolved oxygen. The high dissolved oxygen in the clarifier permitted a certain amount of aerobic metabolism in the clarifier and greatly reduced the time that the bio-

mass was in an anoxic condition.

The total production of solids per pound of BOD added, including underflow waste and effluent solids, was directly related to the solids retention time (SRT) after the soluble substrate (BOD) was converted into biological mass (Figure 8). After an SRT of 6 days, the solids production with the oxygen aeration system was significantly lower than in the conventional system. Indeed, the total solids production decreased to 0.35 pounds of waste solids per pound of BOD added at an SRT of 13 days. The parallel conventional system, operated as step aeration or contact stabilization, increased solids production through an SRT of 9.5 days with a peak solids production of approximately 1 pound of waste solids per pound of BOD compared to 0.5 pounds of solids per pound of BOD for the oxygen system at the same SRT of 9.5 days.

Clarifier Performance

The mixed liquor at the high concentrations employed in the oxygen process settled in a subsidence (zone) settling pattern (14). Discrete interfaces existed between the homogeneous subsiding particles and the decant, and between subsiding particles and the compression zone on the bottom. In the subsidence zone, the relatively uniform particles were nearly homogeneously mixed by the countercurrent turbulence produced by water passing around the solids. The subsidence rate was the controlling factor in the clarifier operation.

After the conversion to center-feed operation in September (Table 2, Period 4), settling problems did not occur even at peak overflows of 1,410 gal./(day)(sq.ft.)-(7.9 ft./hr. rise rate), 1,580 gal./(day)(sq.ft.)-(8.8 ft./hr. rise rate), and 1,940 gal./(day)(sq.ft.)-(10.8 ft./hr. rise rate) in the respective clarifier zones, below the skirt, above the skirt and between the well and weir. The subsidence rate in a 1 liter graduate cylinder

exceeded 8.4 ft./hr. during this period (Figure 9). The blanket level during the peak flow was controllable in the upper zone. The solids loading to the clarifier was 61 lb./(day)(sq. ft.) and the underflow solids concentration was 1.40% solids with 80% volatile material.

In October, the solids level increased with the addition of alum to an average of 8,120 mg./l. The addition of alum increased the solids loading on the clarifier to 88 lb./(day)(sq. ft.). The laboratory subsidence rate decreased to approximately 6.3 ft./hr. (Figure 9). However, by wasting solids ahead of or at peak flows, satisfactory operation was maintained. Again during peak flows, the sludge blanket remained the upper zone of the clarifier.

In November, after reestablishing the bioflocculation lost through the dispersion during alum addition, the laboratory subsidence rate further decreased to a rate of approximately 5.3 sq. ft./hr. (Figure 9). The blanket, as before, remained in the upper zone. Controlled wasting ahead of or during peak flow, as well as frequent reduction of flow, was required to maintain blanket stability. Because of the poor settling qualities after premature addition of alum in December, the mixed liquor solids were allowed to build without chemical addition or without high solids wasting at a steady state flow of 70 gal./min. With maximum laboratory subsidence rates of approximately 7.0 ft./hr. (Figure 10), the sludge blanket in the clarifier did not overflow at 1,280 gal./(day)(sq. ft.)-(7.2 ft./hr. rise rate) in the upper zone. The blanket, however, remained in the upper zone. When the solids level had increased to 5,700 mg./l., the laboratory subsidence rate decreased to 4.0 ft./hr. The pilot plant flow had to be reduced to 50 gal./min. with a clarifier overflow rate of 975 gal./(day)(aq. ft.) (5.1 ft./hr. rise rate) in the upper zone, to prevent uncontrolled solids carryover. By the end of January, the solids were reduced to approximately 3,800 mg./l. by

increased underflow wasting. The laboratory subsidence rate increased to 7.5 ft./hr. and a flow of 60 gal./min. could be maintained in the clarifier. The overflow rate was 1,160 gal./(day)(sq. ft.)-(6.2 ft./hr. rise rate) in the upper zone between the center well and the weir.

The clarifier operation that was observed in September was never regained during the colder winter temperatures even though wide range biological operating modes were used (SRT of 4.7 days to 13.0 days). The subsidence rates in the one liter graduate test during cold temperatures did not regain the 8.4 ft./hr. observed in September. The maximum rate in the cold weather during the month of January was 7.5 ft./hr. Also, the SVI's in the cold weather were higher than in warmer periods (greater than 50 in both operational periods in January).

Oxygen, Mixing, and Recirculation Requirements

Consistently through the operation, vented gas from the fourth stage was less than ten percent of the input oxygen volume. The vented stream was, 50% oxygen. Based upon the influent and exhausted oxygen concentrations, the flow utilization of oxygen in the process was 95%. In Table 5, the accountable oxygen usage consisting of COD removal, nitrification demand, exhaust gas and effluent dissolved oxygen, was compared to the oxygen actually supplied. The difference was probably related to oxygen leakage in the pilot plant operation. Recirculation gas bubbled up and vented through the water-sealed sample ports. Mechanical, instrumental and other miscellaneous oxygen leaks also occurred.

The variable speed 3-h.p. electric motor, three constant speed 3-h.p. electric motors, and the recirculation compressors for mixing were oversized for the system. The propellor-type mixers (3 blades, 14-inch diameter and 11-inch pitch) turned at approximately 190 r.p.m. in each of the four stages. The recirculation rate of the oxygen in the first stage normally was 3-7 cu.ft./min.

and 1-2 cu. ft./min. in each of the last three stages. The total recirculation requirements varied between 0.10 and 0.20 cu. ft./gal. of flow.

Summary

1. A gas-tight biological reactor with independent control of D.O. and mixing, coupled with an aerobic clarifier, produced a good quality secondary effluent in less than 1.66 hrs. average detention time with MLSS up to 6,000 mg./l.

2. The insolublization of biodegradable organics was essentially complete (less then 3 mg./l. of soluble BOD). The carbonaceious BOD removal depended upon the amount of suspended solids in the effluent and therefore on the ability to clarify.

3. The microbial organisms were visually the same as a typical conventional system; however, the percent of volatile solids and their rate of activity were greater at SRT above 6 days.

4. The sludge in the underflow was approximately 1.4% solids without chemical addition and 2.14% with alum addition. The total production of solids was significantly lower than in a parallel conventional system at SRT above 6 days with as little as 0.35 lb. of solids/lb. BOD produced at an SRT of 13 in the oxygen system.

5. In the summer and fall, the solids settled at an average overflow rate of 1,280 gal./(day)(sq. ft.) within the clarifier between the overflow weirs and the center feed well [T,050 gal./(day)(sq. ft.) below the peripheral feed skirt] and up to diurnal peaks of 1,940 gal./(day)(sq. ft.)[T,580 gal./(day)(sq. ft.) below the peripheral feed skirt.] In the winter, the pilot clarifier overflow rates could not exceed 1,160 gal./(day)(sq. ft.) in the zone between the center well and the overflow weir [900 gal./(day)(sq. ft.) below the

increased underflow wasting. The laboratory subsidence rate increased to 7.5 ft./hr. and a flow of 60 gal./min. could be maintained in the clarifier. The overflow rate was 1,160 gal./(day)(sq. ft.)-(6.2 ft./hr. rise rate) in the upper zone between the center well and the weir.

The clarifier operation that was observed in September was never regained during the colder winter temperatures even though wide range biological operating modes were used (SRT of 4.7 days to 13.0 days). The subsidence rates in the one liter graduate test during cold temperatures did not regain the 8.4 ft./hr. observed in September. The maximum rate in the cold weather during the month of January was 7.5 ft./hr. Also, the SVI's in the cold weather were higher than in warmer periods (greater than 50 in both operational periods in January).

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Consistently through the operation, vented gas from the fourth stage was less than ten percent of the input oxygen volume. The vented stream was, 50% oxygen. Based upon the influent and exhausted oxygen concentrations, the flow utilization of oxygen in the process was 95%. In Table 5, the accountable oxygen usage consisting of COD removal, nitrification demand, exhaust gas and effluent dissolved oxygen, was compared to the oxygen actually supplied. The difference was probably related to oxygen leakage in the pilot plant operation. Recirculation gas bubbled up and vented through the water-sealed sample ports. Mechanical, instrumental and other miscellaneous oxygen leaks also occurred.

The variable speed 3-h.p. electric motor, three constant speed 3-h.p. electric motors, and the recirculation compressors for mixing were oversized for the system. The propellor-type mixers (3 blades, 14-inch diameter and 11-inch pitch) turned at approximately 190 r.p.m. in each of the four stages. The recirculation rate of the oxygen in the first stage normally was 3-7 cu.ft./min.

peripheral feed skirt[7]. Larger clarification area is required in the winter on District of Columbia wastewater.

6. Nitrification was achieved in the oxygen aeration system in the summer and fall.

7. Average phosphorus residuals of 1.8 mg./l. as P with alum dosage of 1.4 Al^{+++} to P by weight were achieved.

8. Based upon the influent and exhaust gas flow, over 95% of the oxygen is consistently utilized.

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Table 1. Reactor variables and operating conditions

Operating Period Month Dates	1 June 12-30	2 July	3 August 1-25	4 September	5 October 3-21	6 November 10-30	7 January 1-16	8 January 17-31
Flow Rate (gal./min.)	50-55	80	80	70+	70+*	70+	53	53
Aeration Time (hrs.)	2.00	1.66	1.66	1.95	1.95	1.95	2.50	2.50
Recycle Rate	50%	50%	42%	32%	38%	37%	77%	80%
MLSS (mg./l.)	4140	5180	5250	6000	8120	6350	5300	3940
MLVSS (%)	74%	70%	73%	78%	67%	73%	80%	81%
SRT (days)	7.7	7.3	11.8	10.7	5.5	5.5	13.0	4.7
F/M, $\frac{\text{lb. BOD/day}}{\text{MLVSS}}$	0.333	0.342	0.296	0.304	0.283	0.355	0.275	0.392
Volumetric Loading, $\frac{\text{lb. BOD Demand}}{\text{1,000 cu.ft. tank}}$	57	65	61	79	92	87	65	59
Mixer Power, $\frac{\text{k.w.-hrs.}}{\text{1,000 gal.}}$@	1.27	1.19	0.98	0.92	1.00	1.00	1.18	1.42
Compressor Power, $\frac{\text{k.w.-hrs.}}{\text{1,000 gal.}}$@	0.39	0.28	0.39	0.41	0.35	0.26	0.34	0.32
Temperature, ^{o}F	74-80	78-84	82-85	79-83	70-79	66-69	58-60	58-60

* Alum addition

\+ 2.3:1 diurnal variation

@ Pilot plant equipment efficiency was not determined

Table 2. Clarifier variables and operating conditions

Operating Period	1	2	3	4	5	6	7	8
Month	June	July	August	September	October	November	January	January
Dates	12-30		1-25		3-21	10-30	1-16	17-31
Average Overflow Rate (gal./day)								
at surface	—+	—+	—+	1280	1280	1280	975	975
above feed skirt	750	1210	1210	1050	1050	1050	800	800
below feed skirt	670	1075	1075	940	940	940	710	710
Peak Overflow Rate [gal./(day)(sq.ft.)]								
at surface	—+	—+	—+	1940	1940	1940	975	975
above feed skirt	750	1210	1210	1580	1580	1580	800	800
below feed skirt	670	1075	1075	1410	1410	1410	710	710
Average Solids Loading, lb./(day)(sq.ft.)	37	75	58	61	88	68	55	42
SVI	80	48	50	42	33	48	60	73
Underflow Solids								
% dry solids	1.16	1.34	1.27	1.40	2.14	1.40	1.08	1.00
% volatile	75	70	75	80	65	81	90	80
Waste Solids, lb./million gal.	161	40	144	250	680	230	193	253
Volatile, lb./million gal.	121	28	108	200	441	202	174	202
Effluent Solids, lb./million gal.	296	445	166	204	290	470	197	483
Volatile, lb./million gal.	198	245	113	141	189	342	118	400

+ Peripheral feed - no center feed section

TABLE 3. Organic Removal

Operating Period Month Dates	1 June 12-30	2 July	3 August 1-25	4 September	5 October 3-11	6 November 10-30	7 January 1-16	8 January 17-31
Primary BOD, mg./l.	89	87	89	106	116	131	124	134
Effluent BOD, mg./l.	18	19	12	13	14	27	11	32
Effluent Soluble BOD, mg./l.	-	-	2	2	3	3	3	3
Primary COD, mg./l.	250	244	245	252	284	275	250	256
Effluent COD, mg./l.	45	70	49	51	51	63	59	99
Primary TOC, mg./l.	75	65	77	100	106	91	83	87
Effluent TOC, mg./l.	14	24	15	15	15	21	21	26
Primary Suspended Solids mg./l.	113	101	102	107	120	92	98	100
Effluent Suspended Solids, mg./l.	36	53	28	24	35	56	24	58

Table 4. Nitrogen Balances

Operating Period	1	2	3	4	5	6	7	8
Month	June	July	August	September	October	November	January	January
Dates	12-30		1-25		3-21	10-30	1-16	17-31
Primary TKN, lb./million gal.	198	188	193	206	225	212	215	227
Effluent TKN, lb./million gal.	140	134	85	94	128	166	145	150
Effluent NO_3-NO_2, lb./million gal.	3	15	28	35	22	2	4	3
Waste Sludge TKN, lb./million gal.	13	4	11	19	48	18	17	22

Table 5. Oxygen usage

Operating Period Month Dates	1 June 12-30	2 July	3 August 1-25	4 September	5 October 3-21	6 November 10-30	7 January 1-16	8 January 17-31
Influent COD, lb./million gal.	2080	2030	2040	2090	2370	2290	2080	2140
Effluent COD, lb./million gal.	375	584	408	430	425	525	488	826
Waste COD, lb./million gal.	188+	42+	150+	217+	630+	247	233	336
Removed COD, lb./million gal.	1517	1404	1482	1443	1315	1518	1359	978
Nitrate Demand, lb./million gal.	14	69	128	160	100	18	9	14
Exhaust Oxygen, lb./million gal.	85	54	75	85*	87	85*	65	65
Effluent D.O., lb./million gal.	10*	10*	10	10	10	10*	25	25
Total	1625	1537	1695	1698	1512	1631	1458	1082
Oxygen Supplied, lb./million gal.	1750	1825	1775	1900	1650	±	1700@	1800@

+ COD = 1.4 volatile solids

* Estimate

± Inlet meter malfunctioned

@ Increase sampling and greater losses of O_2 through sample ports

FIGURE 1. Oxygen Aeration System

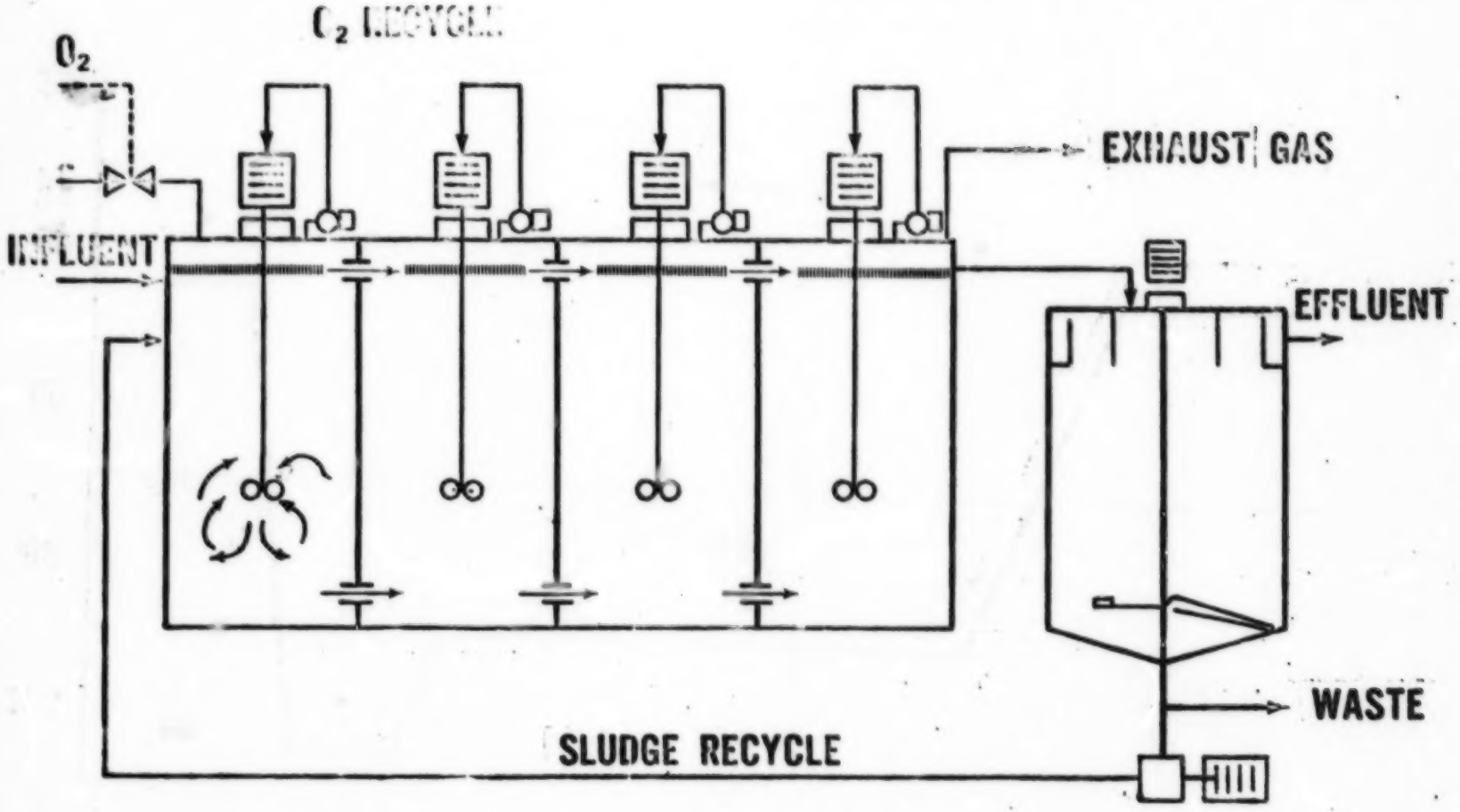
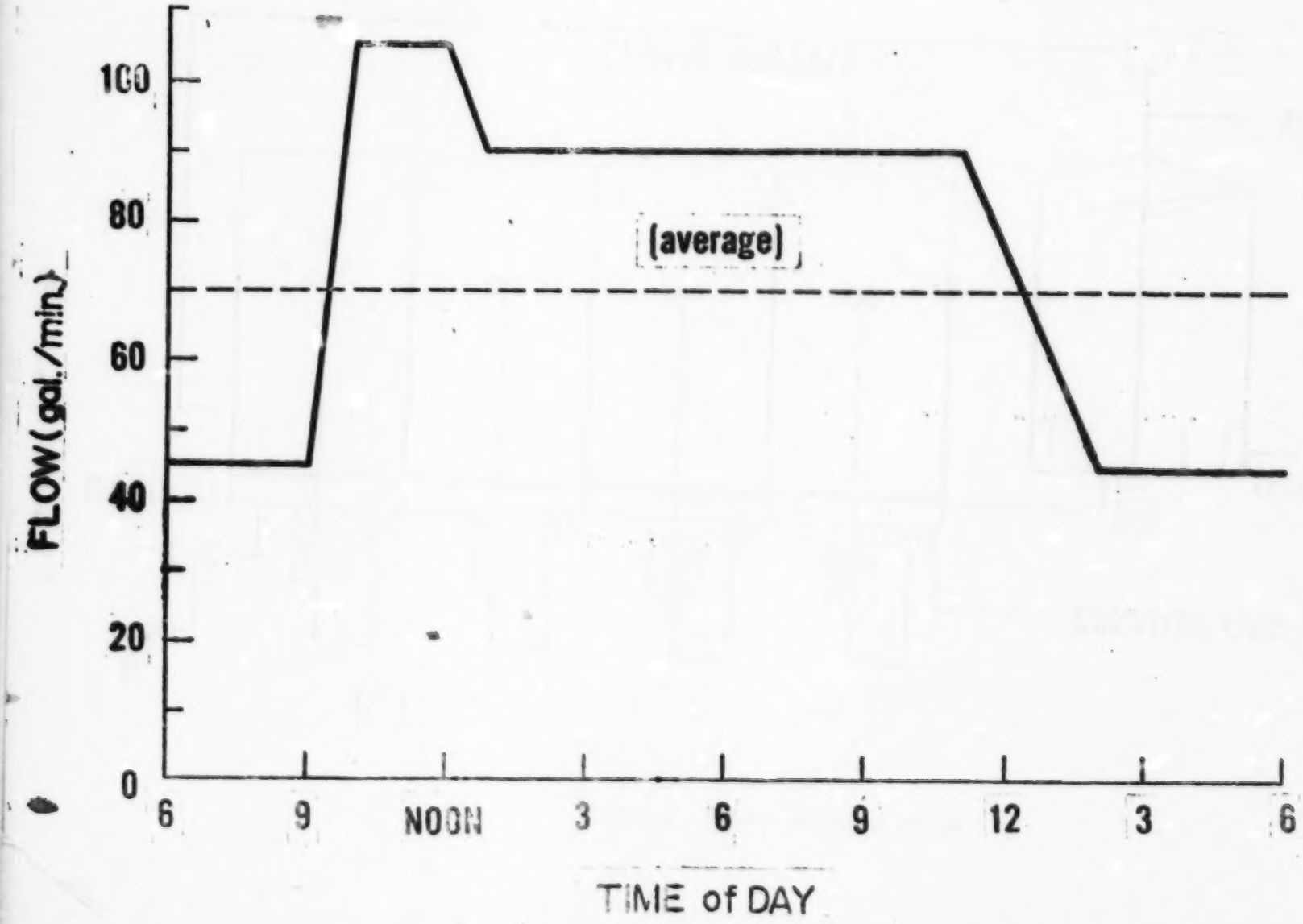


FIGURE 2. Diurnal Flow Variation



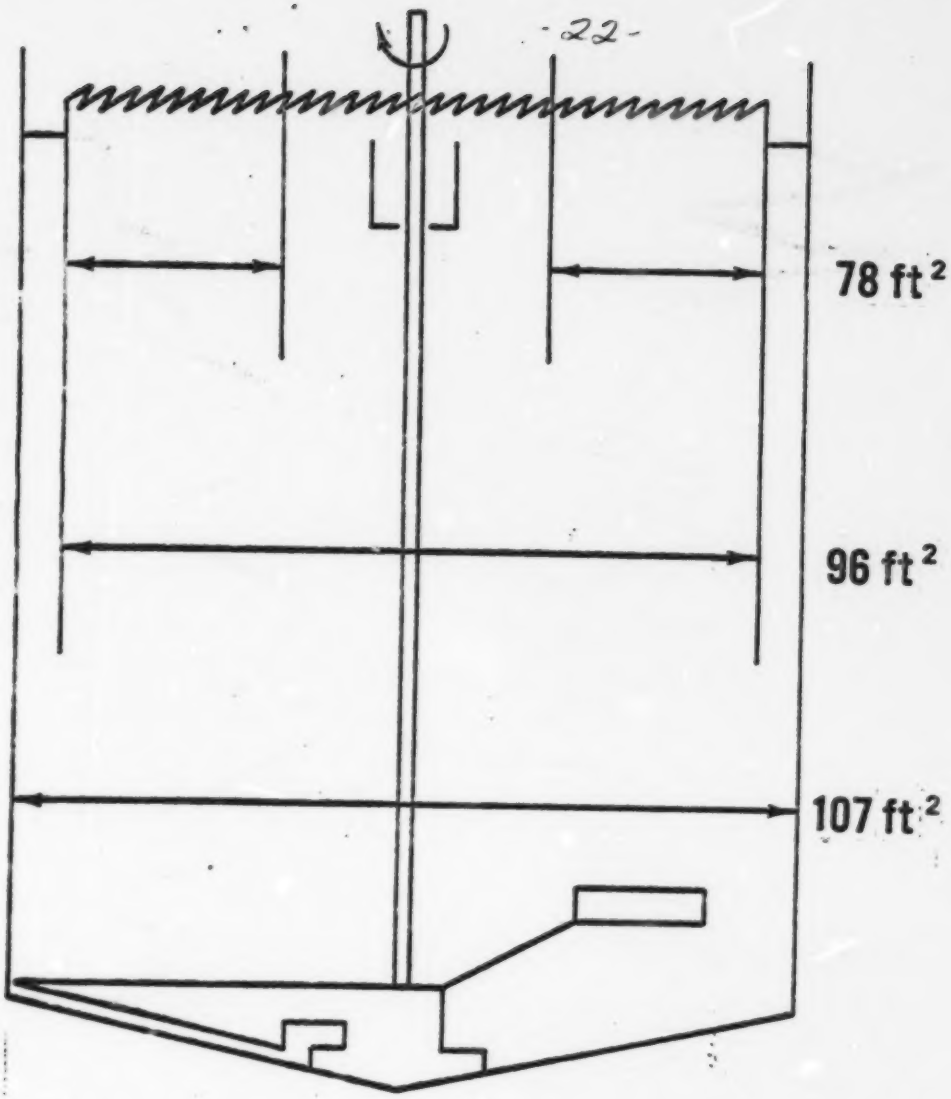


FIGURE 3 - Modified Clarifier with Center Feed

FIGURE 4. Effluent Nitrogen

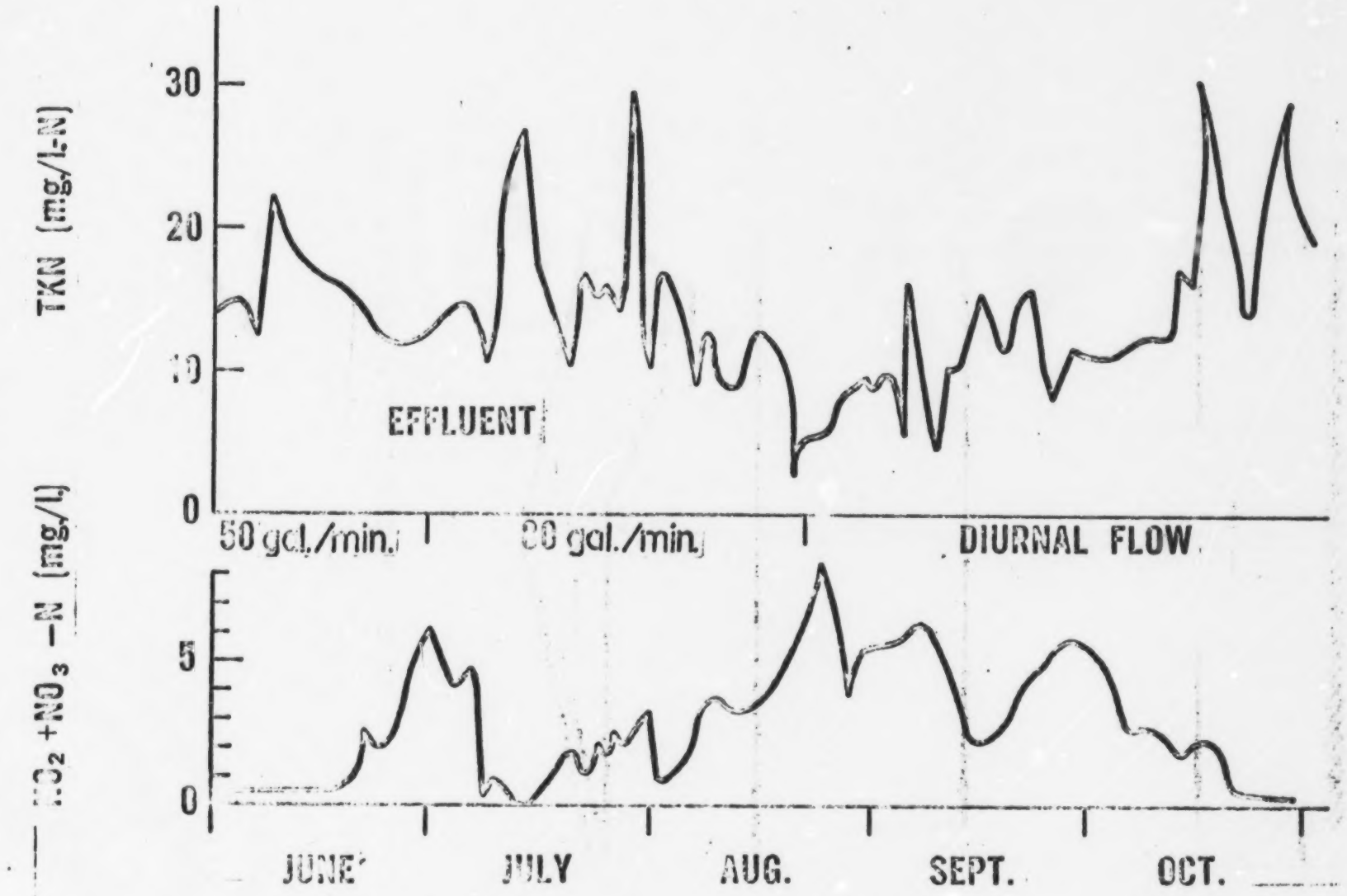


FIGURE 5. Alum Addition

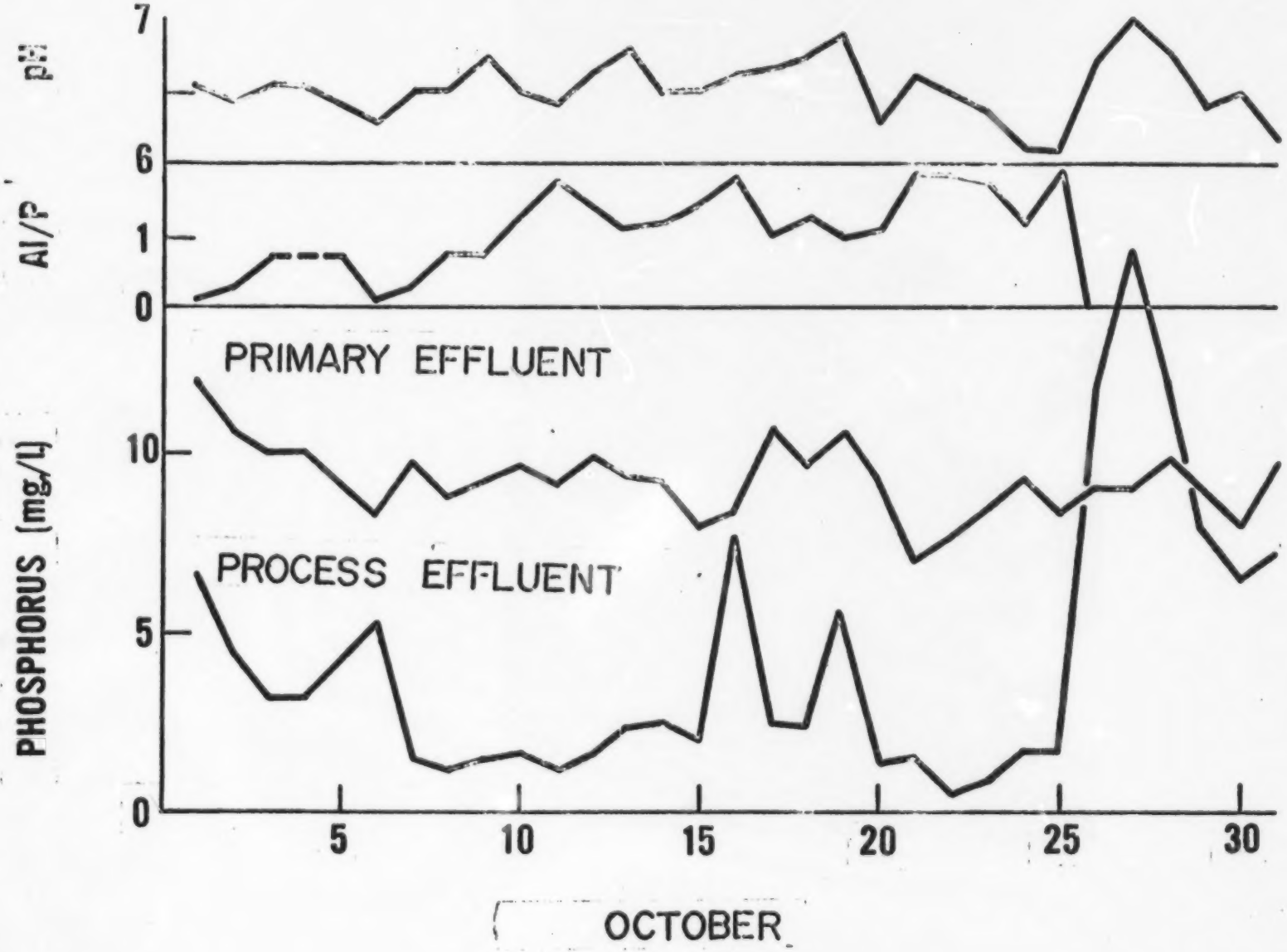


FIGURE 6. Photomicrograph at Sludge Bacteria in Mixed Liquor

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FIGURE 7. Food to Mass Ratio (F/M Versus Solids Retention Time (SRT).

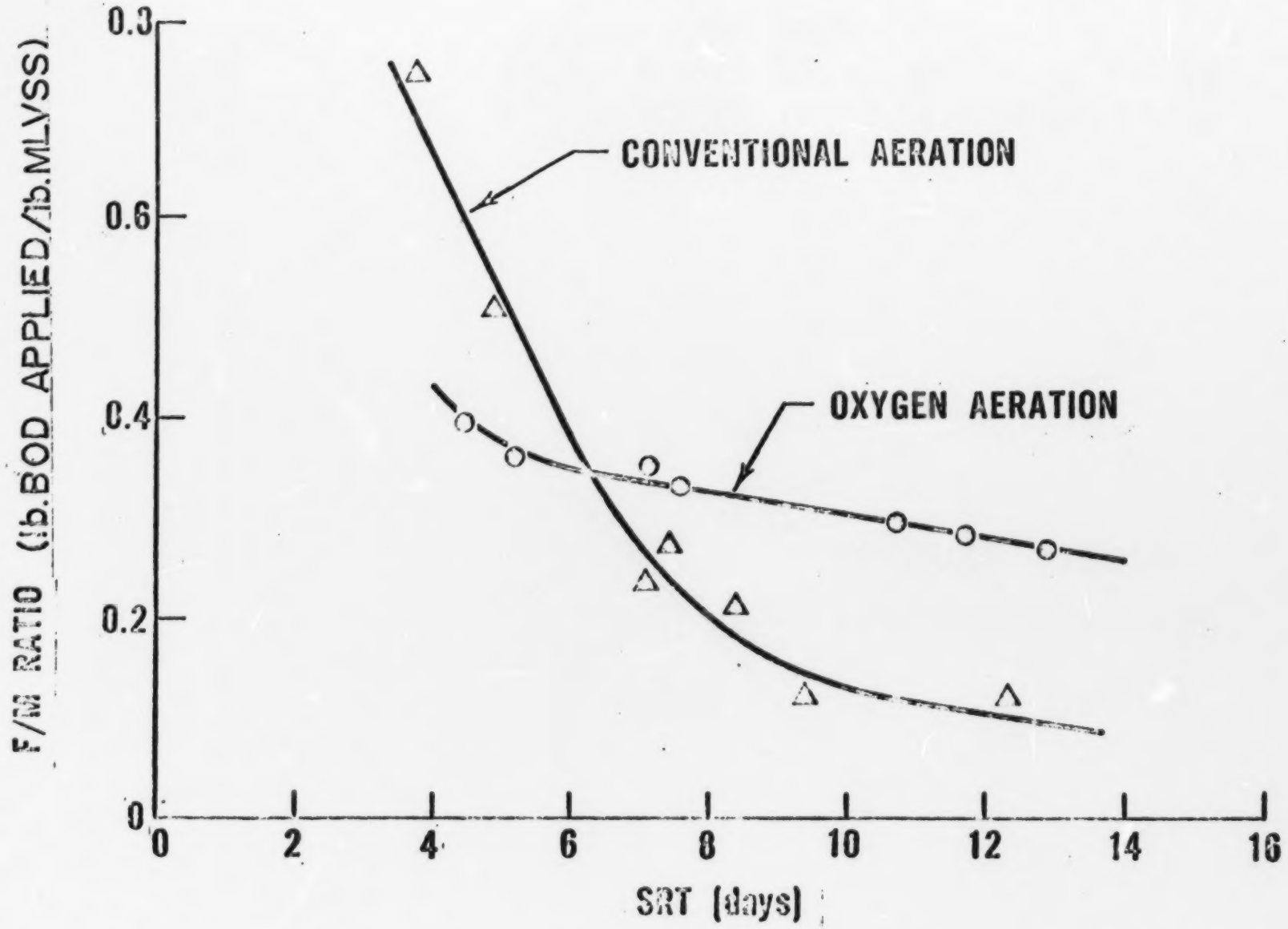


FIGURE 8. Solids Production

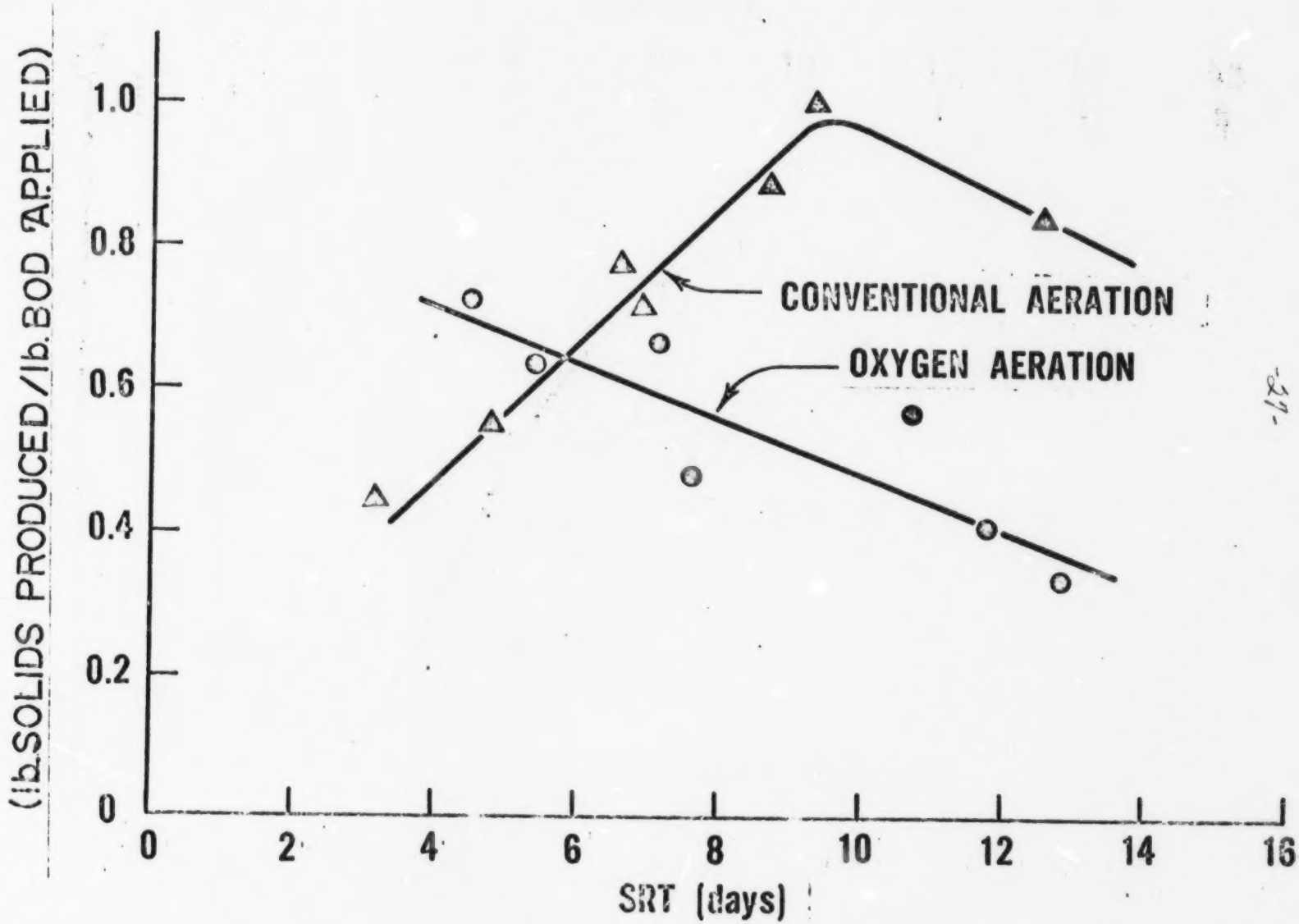


FIGURE 9. 30-Minute Sludge Volume Tests - 9/70, 10/70 and 11/70.

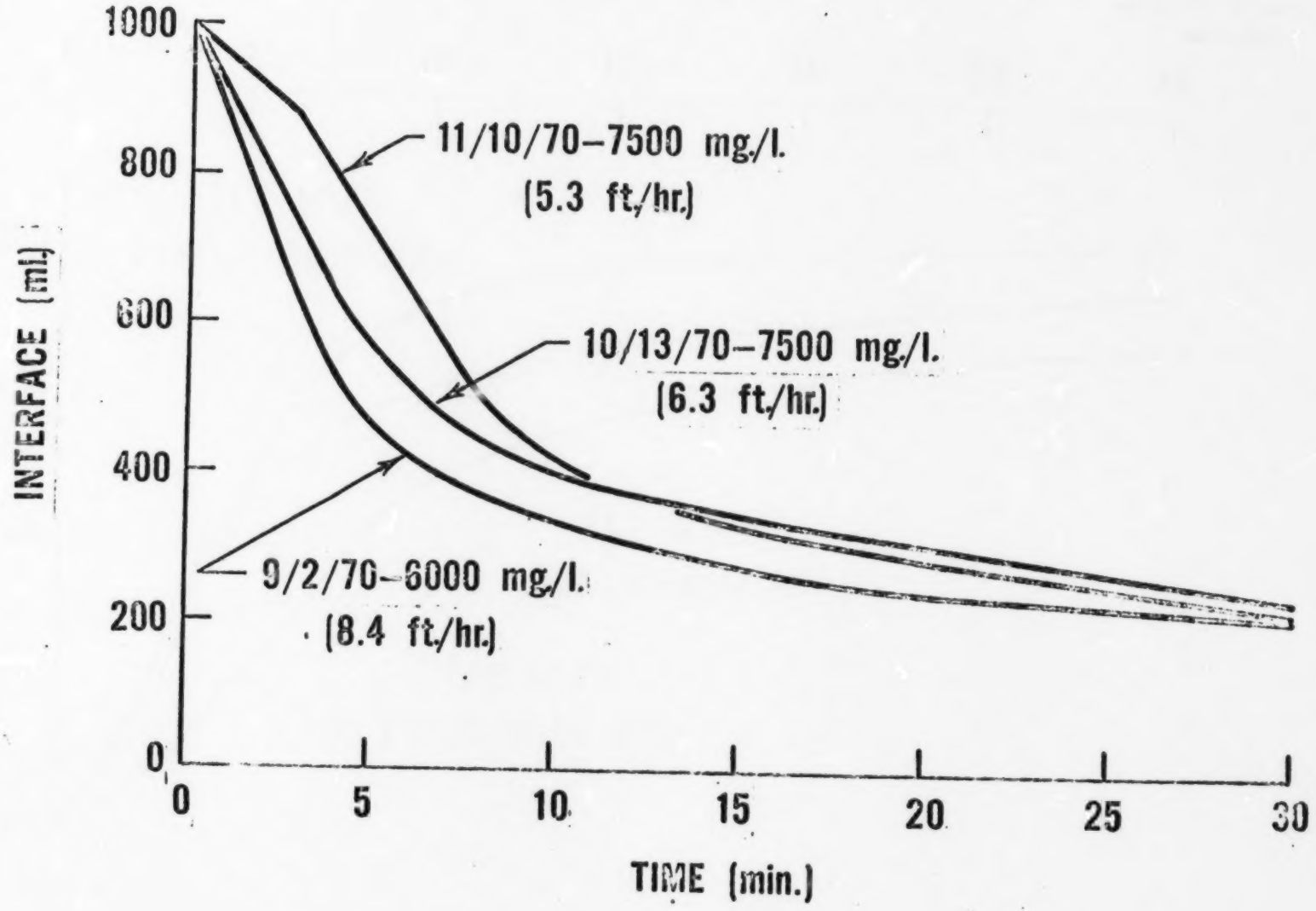
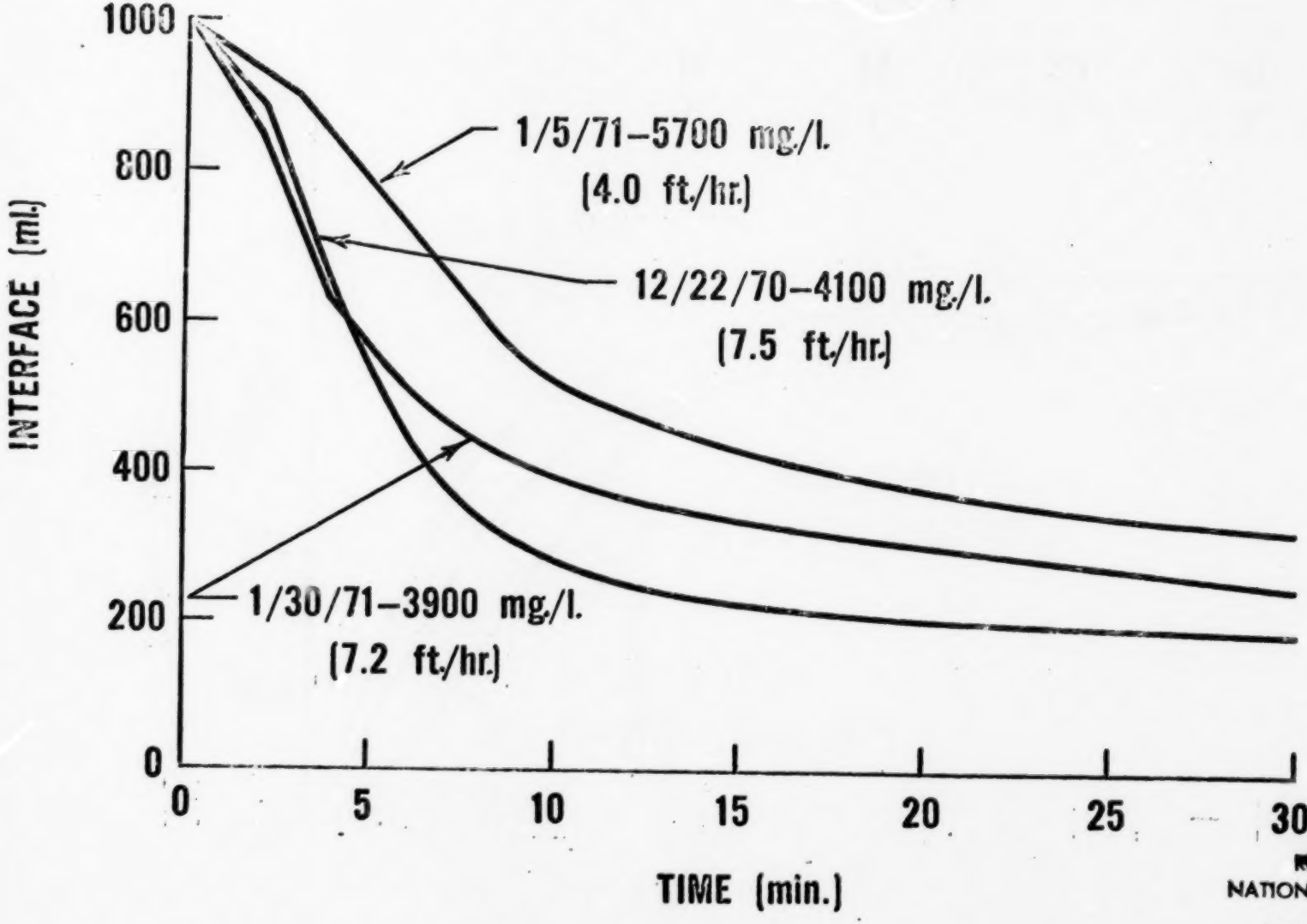


FIGURE 10. 30-Minute Sludge Volume Tests - 12/70 and 1/71



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